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(54) Title: PROCESS FOR PREPARING GAMMA-BUTYROLACTONE, BUTANE-1,4-DIOL AND TETRAHYDROFURAN		
(57) Abstract <p>A process is described for the production of at least one C₄ compound selected from butane-1,4-diol, γ-butyrolactone and tetrahydrofuran, which includes the step of hydrogenation in the vapour phase of maleic anhydride in the presence of a heterogeneous hydrogenation catalyst, which process comprises: (a) contacting a vaporous stream containing maleic anhydride vapour, water vapour, and carbon oxides in an absorption zone with a high boiling organic solvent having a boiling point at atmospheric pressure which is at least about 30 °C higher than that of maleic anhydride thereby to form a solution of maleic anhydride in the high boiling organic solvent; (b) recovering from the absorption zone a waste gas stream; (c) contacting the solution of maleic anhydride in the high boiling solvent with a gaseous stream containing hydrogen thereby to strip maleic anhydride therefrom and to form a vaporous stream comprising hydrogen and maleic anhydride; (d) contacting material of the vaporous stream of step (c) in a hydrogenation zone under hydrogenation conditions in the presence of a heterogeneous hydrogenation catalyst thereby to convert maleic anhydride to at least one C₄ compound selected from butane-1,4-diol, γ-butyrolactone and tetrahydrofuran; and (e) recovering from the hydrogenation zone a product stream containing said at least one C₄ compound.</p>		

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PROCESS FOR PREPARING GAMMA-BUTYROLACTONE, BUTANE-1,4-DIOL AND TETRAHYDROFURAN

This invention relates to the production of butane-1,4-diol, γ -butyrolactone and tetrahydrofuran.

Butane-1,4-diol, together with variable amounts of γ -butyrolactone and tetrahydrofuran, can be produced by hydrogenolysis of maleic anhydride. A major use of butane-1,4-diol is as a feedstock for the plastics industry, particularly for the production of polybutylene terephthalate. It is also used as an intermediate for the production of γ -butyrolactone and of the important solvent, tetrahydrofuran.

Maleic anhydride is normally produced by vapour phase oxidation of a hydrocarbon feedstock, such as benzene, mixed C_4 olefins, or n -butane, in the presence of a partial oxidation catalyst. In the partial oxidation of benzene there is typically used a supported vanadium pentoxide catalyst promoted with MoO_3 and possibly other promoters. The reaction temperature is from about $400^\circ C$ to about $455^\circ C$ and the reaction pressure is from about 1 bar to about 3 bar, while about 4 times the theoretical amount of air is used in order to stay outside the explosive limits. The contact time is about 0.1 s. When the feedstock is a mixed C_4 olefin feedstock, i.e. a mixed butenes feedstock, then the partial oxidation catalyst may be vanadium pentoxide supported on alumina. Typical reaction conditions include use of a temperature of from about $425^\circ C$ to about $485^\circ C$ and a pressure of from about 1.70 bar to about 2.05 bar. The volume ratio of air to butenes may be about 75:1 in order to stay below explosive limits. Alternatively it is possible, according to more modern practice, to design the plant so that satisfactory safe operation can be achieved, despite the fact that the feed mixture of air and butenes is within the flammable limits. In the case of n -butane as feedstock, the

catalyst is typically vanadium pentoxide and the reaction conditions include use of a temperature of from about 350°C to about 450°C and a pressure of from about 1 bar to about 3 bar. The air:n-butane volume ratio may be about 20:1, even
5 though this may be within the flammable limits. One design of reactor for such partial oxidation reactions comprises vertical tubes surrounded by a jacket through which a molten salt is circulated in order to control the reaction temperature.

10 In each case a hot vaporous reaction mixture is recovered from the exit end of the reactor which comprises maleic anhydride vapour, water vapour, carbon oxides, oxygen, nitrogen, and other inert gases, besides organic impurities such as formic acid, acetic acid, acrylic acid, and
15 unconverted hydrocarbon feedstock.

One way of recovering maleic anhydride from such a reaction mixture is to cool it to about 150°C using a steam-producing stream and then to cool it further to about 60°C by cooling it against water in order to condense part of the
20 maleic anhydride, typically about 30% to about 60% of the maleic anhydride present. The remainder of the stream is then scrubbed with water.

Scrubbing with water or with an aqueous solution or slurry is described, for example, in US-A-2638481. Such
25 scrubbing results in production of a solution of maleic acid which is then dehydrated, by distilling with xylene, for example, so as to remove the water and re-form the anhydride. A disadvantage of such a procedure, however, is that an unacceptable proportion of the product remains in the vapour
30 phase. In addition, some of the maleic acid is inevitably isomerised to fumaric acid. The byproduct fumaric acid represents a loss of valuable maleic anhydride and is difficult to recover from the process system since it tends

to form crystalline masses which give rise to process problems.

Because of this isomerisation problem a variety of other anhydrous scrubbing liquids have been proposed. For example, 5 dibutyl phthalate has been proposed as scrubbing liquid in GB-A-727828, GB-A-763339, and GB-A-768551. Use of dibutyl phthalate containing up to 10 weight % phthalic anhydride is suggested in US-A-4118403. US-A-3818680 teaches use of a normally liquid intramolecular carboxylic acid anhydride, 10 such as a branched chain C₁₂₋₁₅-alkenyl substituted succinic anhydride, for absorption of maleic anhydride from the reaction mixture exiting the partial oxidation reactor. Tricresyl phosphate has been proposed for this purpose in FR-A-1125014. Dimethyl terephthalate is suggested for this duty 15 in JP-A-32-8408 and dibutyl maleate in JP-A-35-7460. A high molecular weight wax as scrubbing solvent is taught in US-A-3040059, while US-A-2893924 proposes scrubbing with diphenylpentachloride. Use of an aromatic hydrocarbon solvent having a molecular weight between 150 and 400 and a 20 boiling point above 140°C at a temperature above the dew point of water in the vaporous reaction mixture, for example dibenzylbenzene, is suggested in FR-A-2285386. Absorption of maleic anhydride from the vaporous partial oxidation reaction mixture in dimethylbenzophenone followed by distillation is 25 described in US-A-3850758. Polymethylbenzophenones, at least a portion of which contain at least 3 methyl groups, can be used as liquid absorbent for maleic anhydride according to US-A-4071540. Dialkyl phthalate esters having C₄ to C₈ alkyl groups and a total of 10 to 14 carbon atoms in both alkyl 30 groups are proposed for absorption of maleic anhydride from the reaction mixture in US-A-3891680. An ester of a cycloaliphatic acid, for example dibutyl hexahydrophthalate, is suggested as absorption solvent for maleic anhydride in

ZA-A-80/1247.

It has also been proposed to effect direct condensation of maleic anhydride from the reaction mixture exiting the partial oxidation reactor. However, this procedure is
5 inefficient because an unacceptable proportion of the maleic anhydride remains in the vapour phase.

The maleic anhydride product recovered following condensation or by scrubbing or absorption and distillation can then be subjected to hydrogenation to yield butane-1,4-
10 diol, together with variable amounts of γ -butyrolactone and tetrahydrofuran, as described in US-A-5347021 and EP-B-0373947 the disclosure of which is herein incorporated by reference.

It would be desirable to improve the production of
15 butane-1,4,-diol, γ -butyrolactone and tetrahydrofuran, from maleic anhydride by hydrogenation. In particular it would be desirable to reduce the capital cost of construction of such a plant and also to reduce its running costs, thereby making butane-1,4-diol, γ -butyrolactone and tetrahydrofuran more
20 readily available.

It is accordingly an object of the present invention to improve the production of butane-1,4,-diol, γ -butyrolactone and tetrahydrofuran from maleic anhydride.

According to the present invention there is provided a
25 process for the production of at least one C_4 compound selected from butane-1,4-diol, γ -butyrolactone and tetrahydrofuran, which includes the step of hydrogenation in the vapour phase of maleic anhydride in the presence of a heterogeneous hydrogenation catalyst, which process
30 comprises:

(a) contacting a vaporous stream containing maleic anhydride vapour, water vapour, and carbon oxides in an absorption zone with a high boiling organic solvent having a boiling point at

atmospheric pressure which is at least about 30°C higher than that of maleic anhydride thereby to form a solution of maleic anhydride in the high boiling organic solvent;

(b) recovering from the absorption zone a waste gas stream;

5 (c) contacting the solution of maleic anhydride in the high boiling solvent with a gaseous stream containing hydrogen thereby to strip maleic anhydride therefrom and to form a vaporous stream comprising hydrogen and maleic anhydride;

(d) contacting material of the vaporous stream of step (e)
10 in a hydrogenation zone under hydrogenation conditions in the presence of a heterogeneous hydrogenation catalyst thereby to convert maleic anhydride to at least one C₄ compound selected from butane-1,4-diol, γ -butyrolactone and tetrahydrofuran; and

15 (e) recovering from the hydrogenation zone a product stream containing said at least one C₄ compound.

The vaporous stream of step (a) of the process of the invention is preferably produced by partial oxidation of a hydrocarbon feedstock in the presence of a partial oxidation
20 catalyst using molecular oxygen, typically in the form of air. The hydrocarbon feedstock can be benzene, or a mixed C₄ olefin stream, but is most preferably n-butane. The use of n-butane as hydrocarbon feedstock is currently preferred upon the grounds of cost since it is a cheaper feedstock than
25 benzene or butenes. Hence in the process of the invention the feedstock used for production of the maleic anhydride containing vaporous stream of step (a) is most preferably n-butane and the catalyst is preferably vanadium pentoxide. Typical partial oxidation conditions in this case include use
30 of a temperature of from about 350°C to about 450°C and a pressure of from about 1 bar to about 3 bar, an air to n-butane ratio of from about 15:1 to about 50:1, e.g. about 20:1 and a partial oxidation catalyst comprising vanadium

pentoxide; the contact time is typically from about 0.01 s to about 0.5 s, e.g. about 0.1 s.

Partial oxidation of the hydrocarbon feedstock is conveniently conducted in a reactor which comprises vertical tubes surrounded by a jacket through which a molten salt is circulated in order to control the reaction temperature. The vaporous stream from the partial oxidation reactor can then be cooled by external cooling with boiler feed water to raise steam, and possibly also by further external cooling with cooling water to a temperature in the range of from about 60°C to about 160°C.

In step (a) of the process of the invention the vaporous maleic anhydride stream is preferably contacted with the high boiling solvent at a temperature in the range of from about 60°C to about 160°C, preferably from about 80°C to about 120°C, and at a pressure of from about 1 bar to about 3 bar so as to form a solution comprising maleic anhydride in the high boiling solvent. The contacting can be carried out by bubbling the vaporous stream through a body of the solvent. Alternatively the solvent can be sprayed into the vaporous stream. Countercurrent contacting devices can also be employed wherein the ascending vaporous stream is contacted by a descending stream of solvent in a gas-liquid contacting device, such as a packed scrubber tower or a scrubber tower provided with trays. In this step the solvent will typically be at a lower temperature than the vaporous stream so that the latter is cooled.

In the resulting solution of maleic anhydride in the high boiling solvent the concentration of maleic anhydride in the high boiling solvent may range from about 100 g/l to about 400 g/l.

The high boiling solvent has a boiling point at atmospheric pressure that is at least about 30°C higher than

that of maleic anhydride. The solvent should be selected so that it does not react significantly with maleic anhydride under conditions used in the contacting step (a). Hence it is preferably inert under the scrubbing conditions of step (a)

5 As examples of suitable high boiling solvents there can be mentioned dibutyl phthalate; tricresyl phosphate; dibutyl maleate; a high molecular weight wax; an aromatic hydrocarbon solvent having a molecular weight between 150 and 400 and a boiling point above 140°C, such as dibenzylbenzene; and
10 dialkyl phthalate esters having C₄ to C₈ alkyl groups and a total of 10 to 14 carbon atoms in both alkyl groups. Examples of esters which can be used as the high boiling solvent include di-(C₁ to C₄ alkyl) phthalates, such as dimethyl phthalate, diethyl phthalates, di-*n*- or -*iso*-propyl
15 phthalate, and dibutyl phthalate, di-(C₁ to C₄ alkyl) esters, e.g. dimethyl esters, of other aromatic acids, such as dimethyl 2,3-naphthalene-dicarboxylate, diesters of cyclic aliphatic diacids, such as dimethyl 1,4-cyclohexane-dicarboxylate, and methyl esters of long chain fatty acids
20 containing, for example, from 14 to 30 carbon atoms. Other solvents that can be used include high boiling ethers such as dimethyl ethers of polyethylene glycols of appropriate molecular weight, such as tetraethyleneglycol dimethyl ether.

The high boiling solvent used in step (a) conveniently
25 comprises material resulting from the hydrogen stripping step (c).

Provided that appropriate conditions are adopted in step (a), the gas stream recovered in step (b) of the process of the invention can be essentially free from maleic anhydride.

30 In step (c) of the process of the invention, a gas stream comprising hydrogen is passed through the solution of maleic anhydride.

The hydrogen stripping step is preferably conducted

substantially at or at a pressure slightly higher than the inlet pressure to the hydrogenation zone. The hydrogen stripping step is similarly preferably conducted at substantially the desired inlet temperature to the hydrogenation zone or a little below this temperature, for example from about 5°C to about 20°C below this temperature. Then the temperature can be raised to the desired inlet temperature by admixture of further hot hydrogen-containing gas which has the additional benefit of diluting the vaporous ester-containing stream and thereby ensuring that it is at a temperature above its dew point, preferably at least about 5°C higher than its dew point.

The hydrogenation step is advantageously conducted in the vapour phase, using a heterogeneous hydrogenation catalyst. Typical hydrogenation catalysts include promoted copper-based catalysts, such as a Cu/Zn/Mg/Cr catalyst of the type described in J. Org. Chem 150, pages 177 to 185.

The catalyst particles preferably have a particle size in the range of from about 0.5 mm to about 5 mm. The particles may be of any convenient shape, e.g. spheres, pellets, rings or saddles. When using a fixed bed of catalyst the reactor can be a shell-and-tube reactor, which can be operated substantially isothermally; however, it is preferably an adiabatic reactor. The use of an adiabatic reactor is advantageous since its capital cost is much lower than that of a shell-and-tube reactor and it is generally much easier to charge the reactor with the chosen catalyst.

Hydrogenation is conducted at an elevated temperature of, for example, from about 150°C to about 300°C, more usually from about 180°C to about 280°C, and at a pressure of from about 5 bar to about 100 bar, preferably from about 10 bar to about 70 bar.

From the hydrogenation zone there is recovered a

hydrogenation product mixture which contains, in addition to butane-1,4-diol, also some tetrahydrofuran and γ -butyrolactone. Even if the primary product of interest is butane-1,4-diol, the presence of these minor amounts of tetrahydrofuran and γ -butyrolactone is not disadvantageous since these compounds are important chemicals of commerce and it is accordingly economic to recover them in pure form. If desired, γ -butyrolactone can be recycled to the hydrogenation zone to produce additional butane-1,4-diol.

For further details regarding vapour phase hydrogenation of maleic anhydride reference may be made to a paper by G.L. Castiglioni et al in Erdöl und Kohle - Erdgas - Petrochemie vereinigt mit Brennstoff-Chemie, Bd. 48, Heft 4/5, April/May 1995 at pages 174 to 178 under the heading Wissenschaft & Technik (Science & Technology). Subsequent purification of the resultant crude hydrogenation product mixture can be carried out in a manner analogous to that described in US-A-4584419, WO-A-86/03189, WO-A-88/0937, US-A-4767869, US-A-4945173, US-A-4919765, US-A-5254758, US-A-5310954, and WO-A-91/01960.

In order that the invention may be clearly understood and readily carried into effect a plant for the production of butane-1,4-diol, as well as some γ -butyrolactone and tetrahydrofuran, using a preferred process in accordance with the present invention will now be described, by way of example only, with reference to the accompanying drawing which is a flow diagram of the plant.

Referring to the drawing, n-butane is supplied in line 1 at a pressure of from 1 to 3 bar and at a temperature of 400°C to a partial oxidation plant 2 which is also supplied with air in line 3. Partial oxidation plant 2 is of conventional design and includes a partial oxidation reactor comprising tubes packed with a partial oxidation catalyst

consisting of vanadium pentoxide packed into tubes provided with a jacket through which molten salt can be circulated for the purpose of temperature control. The partial oxidation reactor is operated at an air:n-butane feed ratio of 20:1.

5 A hot vaporous partial oxidation product stream is cooled by external cooling against boiler feed water to raise steam and then against cooling water to reduce its temperature to 138°C. It is recovered from plant 2 in line 4. This contains 2.9% w/w maleic anhydride, 5.8% w/w water, 10 1.3% w/w carbon dioxide, 1.0% w/w carbon monoxide, 0.01% w/w acetic acid, 0.01% w/w acrylic acid, 15.7% w/w oxygen, and the balance essentially comprising nitrogen and other inert gases. It is fed to the bottom of a scrubbing tower 5, up which it passes against a downflowing spray of dimethyl 15 phthalate which is supplied at a temperature of about 68°C from line 6. The scrubbed waste gas stream which contains 0.03% w/w maleic anhydride exits the top of scrubbing tower 5 in vent gas line 7 and is passed to a waste gas burner.

From the bottom of scrubbing tower 5 there is recovered 20 a liquid stream in line 8 which comprises a solution of approximately 15% w/w maleic anhydride and 0.04% w/w acrylic acid in dibutyl phthalate. This is supplied to near the top of a stripping column 9 which is operated at a temperature of 180°C and a pressure of 580 psia (40 bar). Column 9 has a 25 number of distillation trays above the point of injection of the maleic anhydride solution into column 9 so as to reduce carryover of the high boiling solvent dibutyl phthalate in the overhead stream from column 9. The solution of maleic anhydride in dibutyl phthalate flows down stripping column 9 30 against an upflowing stream of hydrogen from line 10. The stripped dibutyl phthalate is recycled from the bottom of stripping column 9 by way of line 6 to the top of scrubbing tower 5. From the top of stripping column 9 there emerges in

line 11 a near saturated vapour mixture stream comprising maleic anhydride in hydrogen, with a hydrogen: maleic anhydride molar ratio of about 400:1. This vapour mixture stream is at a temperature of from about 180°C to about 200°C and at a pressure of about 40 bar. It is diluted with further hot hydrogen from line 12 at a temperature of from about 180°C to about 220°C to yield a vaporous stream with a hydrogen: maleic anhydride molar ratio of about 450:1 and is at least about 5°C above its dew point.

This vaporous mixture passes onwards in line 13 to hydrogenation plant 14 which includes an adiabatic reactor packed with a copper based catalyst (e.g. a promoted copper catalyst) and operated at an inlet temperature of 180°C, an inlet pressure of 565 psia (39 bar), and an exit temperature of 200°C. The maleic anhydride feed rate corresponds to a liquid hourly space velocity of 0.1 h⁻¹. The plant also includes a purification section in which the crude hydrogenation product mixture is distilled in several stages to yield pure butane-1,4-diol in line 15. Lines for separate recovery of γ -butyrolactone and tetrahydrofuran are indicated at 16 and 17 respectively.

CLAIMS:

1. A process for the production of at least one C₄ compound selected from butane-1,4-diol, γ -butyrolactone and tetrahydrofuran, which includes the step of hydrogenation in the vapour phase of maleic anhydride in the presence of a heterogeneous hydrogenation catalyst, which process comprises:
 - (a) contacting a vaporous stream containing maleic anhydride vapour, water vapour, and carbon oxides in an absorption zone with a high boiling organic solvent having a boiling point at atmospheric pressure which is at least about 30°C higher than that of maleic anhydride thereby to form a solution of maleic anhydride in the high boiling organic solvent;
 - (b) recovering from the absorption zone a waste gas stream;
 - (c) contacting the solution of maleic anhydride in the high boiling solvent with a gaseous stream containing hydrogen thereby to strip maleic anhydride therefrom and to form a vaporous stream comprising hydrogen and maleic anhydride;
 - (d) contacting material of the vaporous stream of step (c) in a hydrogenation zone under hydrogenation conditions in the presence of a heterogeneous hydrogenation catalyst thereby to convert maleic anhydride to at least one C₄ compound selected from butane-1,4-diol, γ -butyrolactone and tetrahydrofuran; and
 - (e) recovering from the hydrogenation zone a product stream containing said at least one C₄ compound.
2. A process according to claim 1, in which the vaporous stream of step (a) is produced by partial oxidation of a hydrocarbon feedstock in the presence of a partial oxidation catalyst using molecular oxygen.
3. A process according to claim 2, in which the hydrocarbon feedstock is n-butane.
4. A process according to claim 3, in which the partial

oxidation catalyst comprises vanadium pentoxide and in which the partial oxidation conditions include use of a temperature of from about 350°C to about 450°C, a pressure of from about 1 bar to about 3 bar, an air to n-butane ratio of from about 15:1 to about 50:1 and a contact time of from about 0.01 s to about 0.5 s.

5 10 15 20 25 30 35 40 45 50 55 60 65 70 75 80 85 90 95 100 105 110 115 120 125 130 135 140 145 150 155 160 165 170 175 180 185 190 195 200 205 210 215 220 225 230 235 240 245 250 255 260 265 270 275 280 285 290 295 300 305 310 315 320 325 330 335 340 345 350 355 360 365 370 375 380 385 390 395 400 405 410 415 420 425 430 435 440 445 450 455 460 465 470 475 480 485 490 495 500 505 510 515 520 525 530 535 540 545 550 555 560 565 570 575 580 585 590 595 600 605 610 615 620 625 630 635 640 645 650 655 660 665 670 675 680 685 690 695 700 705 710 715 720 725 730 735 740 745 750 755 760 765 770 775 780 785 790 795 800 805 810 815 820 825 830 835 840 845 850 855 860 865 870 875 880 885 890 895 900 905 910 915 920 925 930 935 940 945 950 955 960 965 970 975 980 985 990 995 1000 1005 1010 1015 1020 1025 1030 1035 1040 1045 1050 1055 1060 1065 1070 1075 1080 1085 1090 1095 1100 1105 1110 1115 1120 1125 1130 1135 1140 1145 1150 1155 1160 1165 1170 1175 1180 1185 1190 1195 1200 1205 1210 1215 1220 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11845 11850 11855 11860 11865 11870 11875 11880 11885 11890 11895 11900 11905 11910 11915 11920 11925 11930 11935 11940 11945 11

which the hydrogen stripping step is conducted at substantially the inlet pressure to the hydrogenation zone.

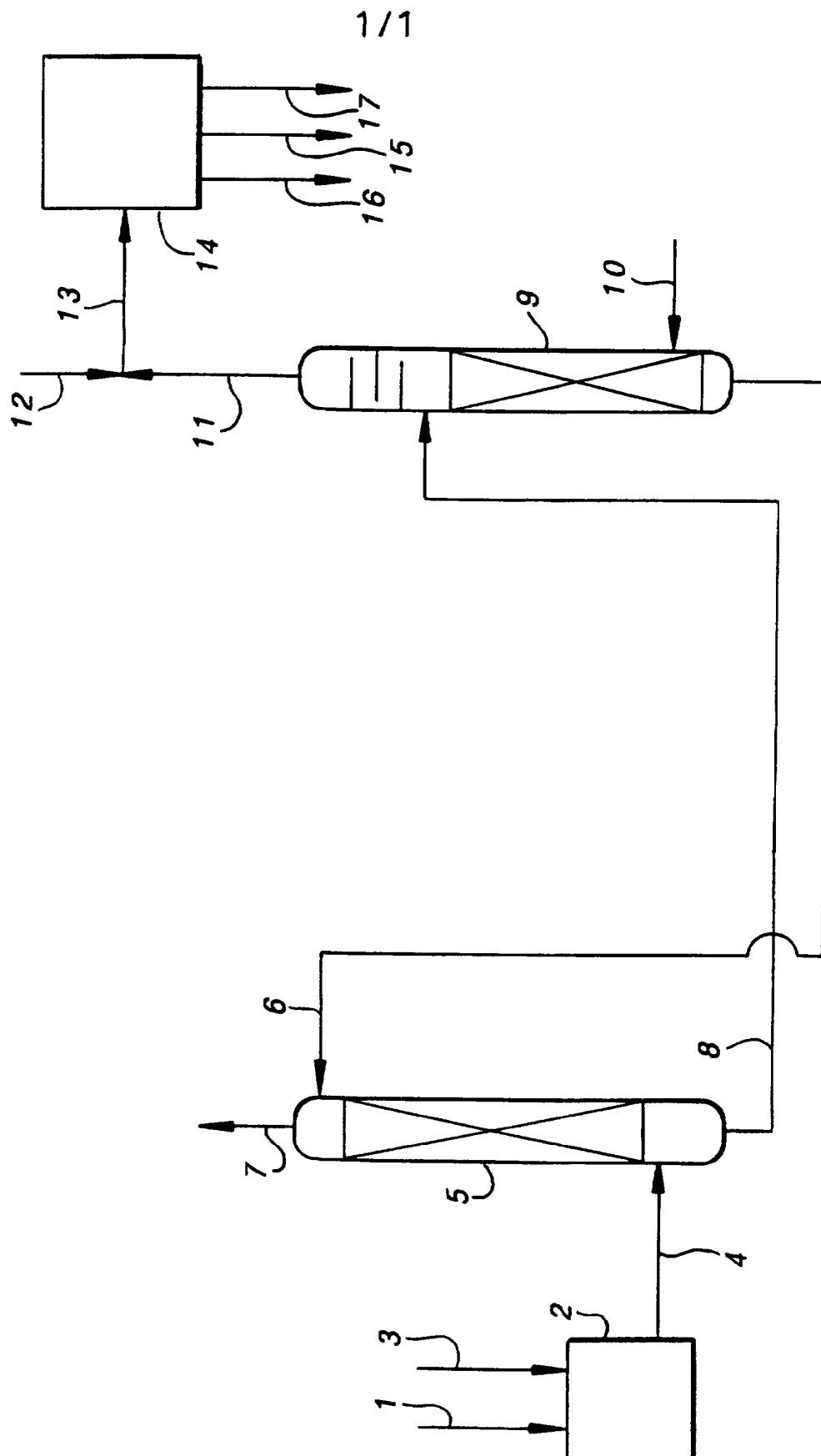
13. A process according to any one of claims 1 to 12, in which the hydrogen stripping step is conducted at a
5 temperature in the range of from the inlet temperature to the hydrogenation zone to about 20°C below the inlet temperature to the hydrogenation zone.

14. A process according to any one of claims 1 to 13, in which the hydrogenation step is conducted in the vapour phase
10 using a promoted copper catalyst at a temperature of from about 150°C to about 300°C and at a pressure of from about 5 bar to about 100 bar.

14. A process according to any one of claims 1 to 13, in which there is recovered from the hydrogenation zone a
15 hydrogenation product mixture which contains, in addition to butane-1,4-diol, also minor amounts of tetrahydrofuran and γ -butyrolactone.

15. A process according to claim 14, in which the hydrogenation product mixture is purified by distillation in
20 one or more stages, including distillation in a "light ends" column to separate overhead the volatile components of the mixture including tetrahydrofuran, and n-butanol.

16. A process according to claim 15, in which the bottoms product from the "light ends" column is further purified by
25 distillation in one or more stages to yield pure butane-1,4-diol.



INTERNATIONAL SEARCH REPORT

International Application No
PCT/GB 97/01286

A. CLASSIFICATION OF SUBJECT MATTER
IPC 6 C07C29/149 C07D307/08

According to International Patent Classification (IPC) or to both national classification and IPC

B. FIELDS SEARCHED

Minimum documentation searched (classification system followed by classification symbols)
IPC 6 C07C C07D

Documentation searched other than minimum documentation to the extent that such documents are included in the fields searched

Electronic data base consulted during the international search (name of data base and, where practical, search terms used)

C. DOCUMENTS CONSIDERED TO BE RELEVANT

Category *	Citation of document, with indication, where appropriate, of the relevant passages	Relevant to claim No.
A	DE 26 42 533 A (BASF AG) 23 March 1978 see the whole document ---	1-16
A	EP 0 373 947 A (TONEN CORP) 20 June 1990 cited in the application see the whole document ---	1-16
A	US 5 347 021 A (TAYLOR PAUL D ET AL) 13 September 1994 cited in the application see the whole document ---	1-16
A	US 3 891 680 A (KATSUMOTO KIYOSHI ET AL) 24 June 1975 cited in the application see the whole document -----	1-16

☐ Further documents are listed in the continuation of box C.

☒ Patent family members are listed in annex.

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- *O* document referring to an oral disclosure, use, exhibition or other means
- *P* document published prior to the international filing date but later than the priority date claimed

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- *&* document member of the same patent family

Date of the actual completion of the international search

18 August 1997

Date of mailing of the international search report

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